Available online at www.sciencedirect.com

SciVerse ScienceDirect

Energy Procedia 23 (2012) 3 – 14

Energy Procedia

Trondheim CCS Conference - 6

Comparison of current and advanced post-combustion CO_2 capture technologies for power plant applications

Miguel A. Gonzalez-Salazar [a]*, Robert J. Perry[b], Ravi-Kumar Vipperla[c], Alvaro Hernandez-Nogales[a], Lars O. Nord[a], Vittorio Michelassi[a], Roger Shisler[b], Vitali Lissianski[b]

[a] *General Electric Global Research, 85748 Garching b. Munich, Germany*
[b] *General Electric Global Research, 1 Research Circle, Niskayuna, NY 12309, US*
[c] *GE Energy, 300 Garlington Road, Greenville, SC 29615, US*

Abstract

Most energy scenarios suggest carbon capture and storage (CCS) from power generation might contribute to reduce the carbon emissions necessary to stabilize the long-term global average atmospheric temperature. GE is actively investigating and developing novel technologies for both capturing and compressing CO_2 from power plants with potential lower energy requirements and environmental impact than state-of-the-art processes. One technology that is currently the focus of significant research effort is phase-changing absorbents for post-combustion capture applications. This investigation compared the performance of phase-changing absorbents to state-of-the-art monoethanolamine (MEA) capture for three different flue gas conditions with CO_2 concentrations ranging from 4 mole% to 13 mole%. Results indicate that depending on the flue gas conditions, the specific equivalent work necessary for operating phase-changing absorbents is expected to be up to 40% lower than for MEA capture. However, as the level of maturity of phase-changing absorbents is certainly lower than MEA capture, higher uncertainty in performance is expected. Besides lower energy requirements, a reduction of up to 6% in specific water cooling load is expected from the phase-changing absorbents compared to MEA capture, in particular for cases with high CO_2 concentrations in the flue gas.

Keywords: CCS; Carbon capture; Post-combustion; Phase-changing absorbents; CO_2 compressi[on]

* Corresponding author. Tel.: +49 (0) 89 55283-549; fax: +49 (0) 89 55283-180.
E-mail address: gonzalez@ge.com

 doi:10.1016/j.egypro.2012.06.057

1. Introduction

Most energy scenarios suggest carbon capture and storage (CCS) from power generation might contribute to reduce the carbon emissions necessary to stabilize the long-term global average atmospheric temperature. While renewables would likely keep growing worldwide in the future, CCS from power plants would still be required to respond to an increasing energy demand while meeting emission targets. CCS technologies mainly address coal-fired power generation, partly because it offers the potential to reduce over 40% of the energy-related anthropogenic greenhouse gas emissions. In addition, applying CCS to other power plants combusting carbon containing fuels might offer even further potential to reduce emissions.

GE is actively investigating and developing novel technologies for both capturing and compressing CO_2 from power plants with potential lower energy requirements and environmental impact than state-of-the-art processes. One technology that is currently the focus of significant research effort is phase-changing absorbents for post-combustion applications.

This investigation compared the performance of phase-changing absorbents to state-of-the-art monoethanolamine (MEA) capture for three different flue gas conditions with CO_2 concentrations ranging from 4 mole% to 13 mole%. Evaluated applications included retrofit and greenfield power plants. While MEA is considered a mature and near commercial technology that might be employed in retrofit and greenfield applications, phase-changing absorbent is considered a next generation capture technology and its performance was evaluated only for greenfield applications. With regard to CO_2 compression, an integrally geared compression train with supercritical pumping was evaluated, as this solution proved to be the least energy intensive for a wide operational range. Aspen Plus® and Thermoflex© were used to simulate the performance of both technologies for the different study cases. As the energy requirements for the two capture technologies varied qualitatively, the concept of specific equivalent work (MJ/kg-CO_2) was used for comparing the performance of the capture technologies. Finally, the specific water cooling load (MJ/kg-CO_2) was also estimated.

2. Approach

Most studies in literature comparing the performance of CO_2 capture technologies for power plants applications used two different methodologies. On one hand, some studies included very detailed models of the power plant and its interaction with the capture unit [1]-[3]. On the other hand, some other studies did not include any detail of the power plant and focused only on the capture unit [4]-[6]. In this study, priority was given to understand the performance of the capture and compression technologies for generic flue gas conditions, rather than the performance of specific power plants with CCS. Thus, the performance of both phase-changing absorbents and MEA was estimated at 90% capture for three different flue gas conditions with CO_2 concentrations ranging from 4 mole% to 13 mole% (see Table 1). These selected flue gas conditions are representative for large scale power plants fuelled with fuels ranging from natural gas to coal.

Table 1. Flue gas conditions for study cases.

	Case 1	Case 2	Case 3
CO_2 concentration (mole%)	4	8	13
H_2O concentration (mole%)	9	10	17
N_2 concentration (mole%)	75	78	67
O_2 concentration (mole%)	12	4	2
Temperature (°C)	57	57	57
Pressure (bar)	1,014	1,014	1,014
Flow rate (kg/s)	650	650	650

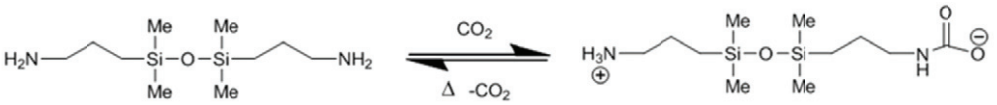
To estimate the mean and variability in performance for both capture technologies a Six Sigma™ methodology was followed. The methodology can be summarized in the following steps:

1. Identify key performance indicators (dependent variables) to compare both technologies, in this case the specific equivalent work and the specific water cooling load are used.
2. Identify the key design parameters (independent variables) for both technologies that influence the performance indicators.
3. Define mean values and standard deviation for the identified independent variables.
4. Design experiment to evaluate the effect of the independent variables and their interactions on the dependent variables.
5. Run the experiment using software process simulation.
6. Build transfer functions that define the dependent variables in terms of the independent variables.
7. Quantify the mean and standard deviation of each of the performance indicators.

A more detailed description of phase-changing absorption and MEA capture as well as the design process is presented in next sections.

3. Phase-changing absorbent

The phase-changing absorbent is based on an aminosilicone material (1,3-bis(3-aminopropyl)-1,1,3,3-tetramethyldisiloxane) [GAP-0] as shown in Equation 1. This substrate is a high boiling (265 °C), low viscosity (4 cP @ 25 °C) liquid that readily forms a solid on contact with CO_2. This solid carbamate salt is readily converted back into the starting aminosilicone on thermal treatment.



GAP-0 GAP-0 Carbamate (1)

The aminosilicone offers a number of advantages over the traditional aqueous MEA system. This material is more thermally stable, has a lower vapor pressure, a lower heat capacity and a substantially lower freezing point than MEA. Additionally, the aminosilicone sorbent is used in a non-aqueous based

process which significantly reduces the amount of energy required to heat and condense water found in the MEA system. In this novel procedure, the neat GAP-0 liquid readily reacts with the CO_2 in the flue gas to generate a solid. While some water will be present in the flue gas, experiments have indicated that GAP-0 exposed to wet CO_2 still maintains its friable solid character with no loss in capture capacity. A preliminary process model was developed for the CO_2 separation unit to calculate the mass and energy balances and the system performance. The process model was calibrated with experimental data. The model accounts for capture of CO_2 by GAP-0 sorbent, heat input needed to desorb the CO_2 and the heat input needed for sensible heating of the sorbent.

3.1. Process description

The general process envisioned consists of four unit operations (see Fig. 1). The first is CO_2 absorption and formation of the solid. This may occur in a spray tower configuration (absorber) wherein fine droplets of lean GAP-0 sorbent are sprayed into the CO_2-rich flue gas stream at approximately 50°C. The carbamate particles that are formed are isolated and collected in a second operation which may be a cyclone-type separator. The third action is transport of the solid from a low pressure (~ 1 bar) regime to the high pressure desorber which may be between 5-20 bar. This may be accomplished by the use of a Posimetric Pump®[1]. The rich sorbent from the absorber is fed to the rich-lean heat exchanger and heated to temperatures of 100-115°C before being fed to the desorber for separation of the absorbed CO_2. The final step is thermal desorption of the CO_2 from the sorbent at ~125 °C under pressure. Steam is supplied to the desorber to provide heat, which releases CO_2 from the rich sorbent. The hot vapor from the top of the desorber consisting of CO_2 is cooled in a heat exchanger utilizing water. The stream then flows to a separator where the vapor and entrained liquid are separated. The CO_2 gas is removed from the separator and then delivered to the CO_2 compressor. The liquid from the bottom of the separator is returned back to the desorber. The lean solvent that is reformed is returned to the absorber unit for further reaction with CO_2. Dissociation of the CO_2 at elevated pressures will reduce compression costs of making supercritical CO_2 for transportation and storage.

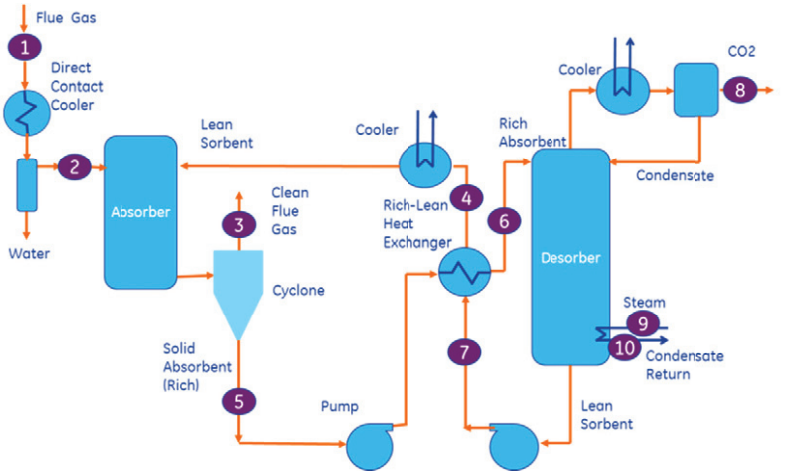


Fig. 1 CO_2 Island process flow diagram

[1] Posimetric is a registered trademark of the General Electric Company.

3.2. Key system assumptions

The energy needed by the CO_2 separator is provided by extraction of steam from the power plant. The steam will pass through the desorber reboiler, and must have a condensation temperature as high as the temperature in the desorber. The system has four process variables that dominate the performance: absorber temperature, desorber temperature, desorber pressure, and rich/lean heat exchanger approach temperature. The system model accounts for the major energy penalties for CO_2 separation, and they include the energy required:

1. For vaporization of water.
2. For desorbing the carbon dioxide (i.e. reaction energy).
3. For sensible heating of the sorbent.

The model also accounts for CO_2 compression energy and auxiliary loads. The sorbent rich loading is defined as the weight percentage of CO_2 in the rich sorbent leaving the absorber column. The sorbent lean loading is defined as the weight percentage of CO_2 in the lean sorbent leaving the desorber column. The sorbent net loading is defined as the difference between the rich loading and the lean loading and is obtained from lab-scale experiments. The lab-scale isotherm data indicate that sorbent net loading of 8% is achievable with GAP-0. The key assumptions for the CO_2 separation unit utilizing the GAP-0 sorbent are listed in Table 2.

Table 2. Parameters used in the baseline (GAP-0).

Parameter	Value
Temperature of flue gas after direct contact cooler (^oC)	32
Absorber temperature (^oC)	49
Absorber pressure (bar)	1,03
Desorber temperature (^oC)	127
Desorber pressure (bar)	13,8
Rich-lean heat exchanger temperature approach (^oC)	5,5

The GAP-0 sorbent utilizes less energy than the MEA sorbent due to lower water in the sorbent mixture and a low specific heat of the sorbent.

Low water in the sorbent mixture

The model accounts for absorption of water in the flue gas by the MEA sorbent and the vaporization of water in the desorber column. The baseline MEA sorbent concentrations are limited to 20-30% and the remaining is water due to viscosity and corrosion issues. The water in the sorbent necessitates significant amount of energy due to sensible heat as well as vaporization of the water.

Low specific heat of the sorbent

The specific heat of GAP-0 is 2,3 kJ/kg-°C while the specific heat of MEA is 3,73 kJ/kg-°C. The lower specific heat for GAP-0 improves the energy efficiency of the process.

3.3. Design of Experiment (DOE)

The parameters that were varied for the design of experiment and the mean and standard deviation values are given below (see Table 3).

Table 3. Design of Experiment (DOE) for GAP-0.

	Mean	Standard deviation
Desorption pressure (bar)	15,85	3,45
Net loading (%)	7,3	2
Heat of reaction (kJ/kg)	1768	186
Rich lean heat exchanger temperature approach (°C)	11	5,5

4. Monoethanolamine (MEA) capture

The investigated capture configuration consists of an absorber with intercooler and a stripper, which is a state-of-the-art and demonstrated technology. The selected solvent is 30%wt MEA, which is characterized by high energy requirements, thermal degradation above 125°C and corrosion. In the absorber column the CO_2 contained in the flue gas reacts with the aqueous MEA. The configuration with absorber intercooler was chosen to improve the exothermic absorption reaction. This improvement is especially significant at high CO_2 concentrations (higher than 8%) and high lean loadings (higher than 0,28 mol CO_2/mol MEA). In the desorber, the reaction is reversed and the absorbed CO_2 is heated up and stripped from the solution. Finally, purified CO_2 is sent for compression while the regenerated solvent is cooled down and sent back to the absorber. The plant was modeled with Aspen Plus® 7.1 and Aspen RateSep™ (rate-based distillation). Fig. 2 shows the capture process flow diagram.

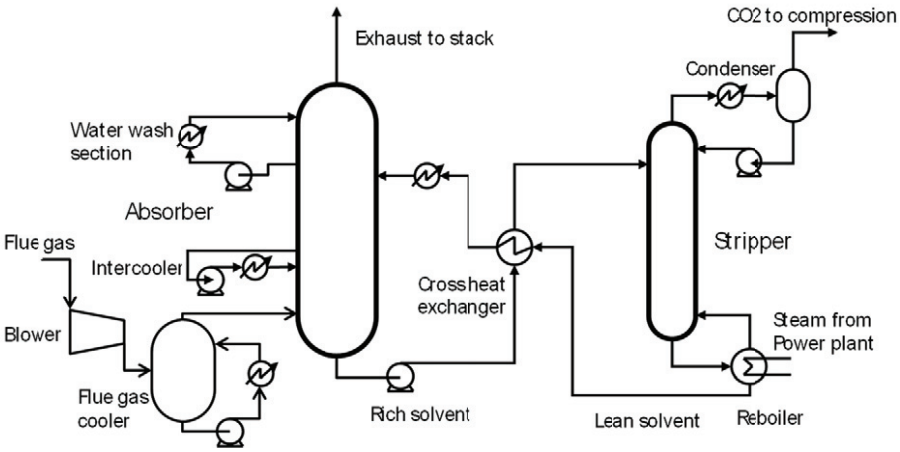


Fig. 2. MEA capture process

One simple train is used to process all the exhaust gases from the power plant. In spite of previous studies that suggest the use of two absorber trains and one stripper [7], [8], this study follows the approach of Fluor that proposes the construction of bigger absorber diameters to reduce capital cost [9]. The flue gas blower is set upstream of the flash cooler. Compared to the downstream position, the blower duty is higher but the absorber of flue gas inlet temperature is reduced, enhancing the absorption process. Blower outlet pressure is automatically designed to overcome the pressure drops in the flue gas cooler and the absorber. Table 4 shows the fixed parameters used in the simulation. The solvent rate is automatically varied to achieve 90% capture in the absorber. The stripping process is designed to desorb the same quantity of CO_2 that was captured in the absorption column. Therefore, the solvent lean loading from the stripper outlet is the same as the absorber solvent inlet.

The stripper pressure is set to 1.9 bar to avoid possible MEA degradation (125°C). At this pressure the reboiler temperature varies from 114°C and 120°C, depending on the loading. As the stripper pressure increases, the driving forces to strip the CO_2 are enhanced, reducing the reboiler duty and the compression work. However, a higher stripper working pressure needs better quality steam to feed the reboiler, which increases the energy penalty. After the capture process, the CO_2 is compressed to supercritical conditions (150 bar) by using a 6-stage integrally geared compressor train with intercooling at 45°C. Fig. 3 shows the CO_2 compression process flow diagram.

Table 4. Parameters used in the baseline (MEA).

Parameter	Value
Capture ratio	90%
Stripper pressure (bar)	1,9
Stripper height (m)	10
Columns diameter condition	80% flooding
Temperature of flue gas and solvent at absorber inlet	40°C
Stripper condenser temperature (°C)	40°C
Packing type	FLEXIPAC® 1Y

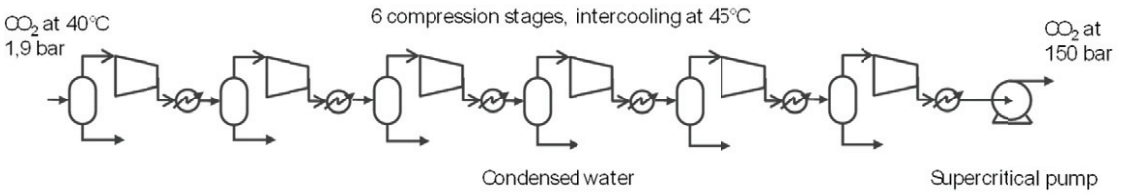


Fig. 3. CO_2 capture process flow diagram

4.1. Design of Experiment (DOE)

In order to evaluate the behavior of the capture plant under different flue gas conditions, three variables were selected: lean loading, absorber height and heat exchanger temperature approach. The lean loading and the heat exchanger temperature approach affect the energy requirements in the stripper, in particular the sensible and the latent heat. The absorber height affects significantly the absorption capacity and water cooling load in the absorber and just slightly the energy requirement in the stripper. Table 5 shows the selected parameters for the design of experiments. For the sake of brevity not all steps of the Six Sigma™ methodology are shown.

Lean loading

The lean loading is defined as the molar ratio of CO_2 to MEA in the absorber inlet solvent stream. A low lean loading means a high capacity of the solvent to absorb CO_2, but also a lower CO_2 partial pressure at the bottom of the stripper which means a higher amount of energy to desorb CO_2. Although other studies consider lean loading levels higher than 0,3 mol CO_2/mol MEA [4], [10], the market prefers lower loading levels to reduce the absorber capital cost. Based on previous experience and data found in the literature [7], [8], the selected most likely values for the lean loading are 0,25, 0,27 and 0,29 for Case 1, 2 and 3 respectively, see Table 5. It is important to note that while these are most likely values, optimizing the lean loading for each case was not in the scope of this work.

Absorber height

The absorber height was varied for the three flue gas conditions. As the CO_2 concentration of the flue gas increases, higher solvent rate is needed to achieve 90% capture rate. While the diameter of the columns is automatically designed to achieve 80% flooding, it is still necessary to adapt the height to the increasing solvent rate for the different flue gas conditions. Based on previous experience and data found in the literature [7], [8], the most likely absorber heights are 15, 20 and 25 m for Case 1, 2 and 3 respectively.

Heat exchanger temperature approach

The cold side temperature approach of the heat exchanger considerably affects the sensible heat requirements in the reboiler duty. Recent papers [10] show the possibility of using 5°C instead of 10°C to improve the performance of the plant. This reduction leads to a strong increase in the capital cost. The suitability of using a smaller or higher temperature approach will be determined by the business plan. In this investigation the selected most likely value for the heat exchanger temperature approach is 9°C and that agrees with another study [11].

Table 5. Design of Experiment (DOE) for MEA

	Case 1	Mean Case 2	Case 3	Standard deviation
Absorber height (m)	15	20	25	1
Lean loading (mol CO_2 / mol MEA)	0,25	0,27	0,29	0,005
Heat exchanger approach (°C)	9	9	9	0,8

5. Equivalent work

As the energy requirements for the phase-changing absorbent and MEA varied qualitatively, the concept of specific equivalent work was used for comparing the performance of the capture technologies. The specific equivalent work has been used in the literature to compare the overall energy requirements (heating, electricity) of different process configurations, capture technologies or solvents [4], [6], [12]. Rochelle et al. define the specific equivalent work as the sum of the electric power consumed in the process (CO_2 compressor, pumps, flue gas blower, others) and the work that otherwise could be generated with the steam condensing in the reboiler, assuming a 75% Carnot efficiency (see Equation 2).

$$W_{eq} = \frac{0{,}75\ Q\left(\frac{T_{Heating}-T_{Sink}}{T_{Heating}}\right)+\sum W_{pump/compressor}}{\dot{m}_{CO2}} \quad (2)$$

While the specific equivalent work as defined above might be useful to compare capture technologies without the need for specifying details of the power plant, it does not fully describe the overall energy penalty. In particular the first term of the equation, defined as the work that could be generated with the steam condensing in the reboiler ($T_{Heating}$), assumes that the steam extracted from the power plant is saturated and that the heating process is isothermal. However, extraction steam at the specific pressure required in the reboiler (~3 bar) rarely occurs in most of today's steam power plants and when it occurs is in superheated condition. This means that the actual extraction temperature is much higher than the saturation temperature required in the reboiler (max. 125°C for MEA to avoid solvent degradation) and therefore the extracted steam should be desuperheated. This desuperheating effect is though not described in Equation 2.

An alternative to account for the desuperheating effect in the specific equivalent work is suggested here (see Fig. 4). The approach of converting the heating requirements of the capture plant into specific equivalent work is accomplished in two steps. In the first step, the needed steam flow is calculated in Thermoflex© based on the heat requirements of the desorption process (Q) and the conditions of the extraction steam. It is assumed that the reboiler has a pinch temperature of 10°C and therefore the required steam temperature should be 10°C higher than the reboiler temperature. As such, the conditions of the steam required for the desorption process with MEA are 2,7 bar/130°C, 2,6 bar/128,8°C and 2,5 bar/127,6°C for cases 1, 2 and 3 respectively. For the phase-changing absorbent the conditions of the required steam are 2,47 bar/127°C. Regarding retrofit and greenfield applications, it is assumed that the extraction steam conditions in both cases are different. For retrofit applications extraction steam conditions are assumed to be those of state-of-the-art supercritical steam power plants, i.e. 5 bar / 291°C (Case 11 from DOE/NETL report [1]). For greenfield applications it is assumed that future steam power plants will be designed to have steam extraction close to the conditions required for the desorption process, i.e. 3,1 bar/135°C. Note that as the pressure and temperature of the available steam are higher than required, a throttle valve and a desuperheater are used to ensure the right conditions. Throttling and desuperheating the extraction steam have been commonly used in the CCS literature [1]-[3].

Once the amount of extraction steam is estimated, the second step is calculating the equivalent power that could be otherwise generated. For this purpose a simplified process layout including a low-pressure steam turbine, a condenser and water pumps was built in Thermoflex©. A condenser pressure of 0,069 bar (1 psia) and a dry step efficiency of 90% for the steam turbine are assumed. Although the described

two-steps approach accounts for the desuperheating effect in the specific equivalent work, it does not account for the effect of running the LP turbine in off-design for retrofit cases (effects include mainly pressure drop at the inlet and reduced efficiency).

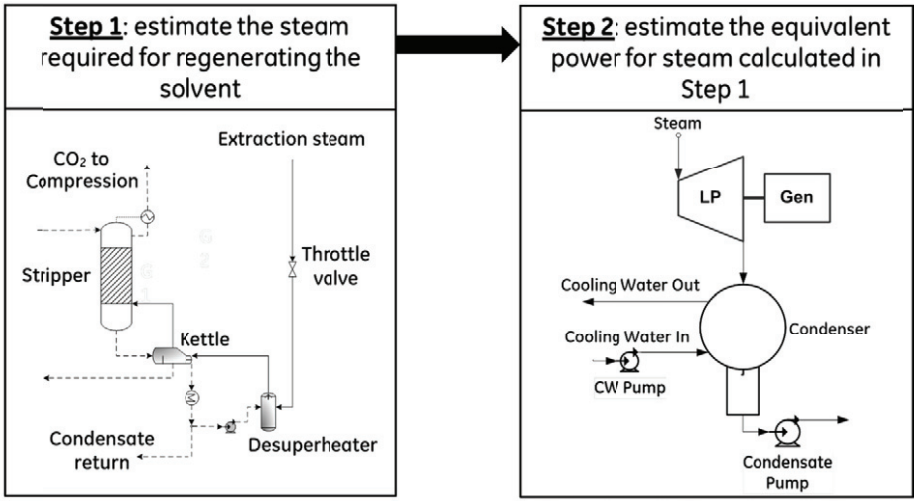


Fig. 4. Two-steps approach for calculating the specific equivalent work.

The modified expression for the specific equivalent work is shown in following equation:

$$W_{eq} = \frac{W_{Eq.\ Turbine} + W_{Pumps} + W_{Blower} + W_{Compressor}}{\dot{m}_{CO2}} \quad (3)$$

Where $W_{Eq.Turbine}$ is the equivalent power calculated in the two-steps approach described above. In addition to the specific equivalent work, the specific water cooling load is also estimated. The specific water cooling load is defined as follows:

$$WCD = \frac{\sum Cooling\ Load_{intercoolers,flue\ gas\ cooler,etc}}{\dot{m}_{CO2}} \quad (4)$$

6. Results and discussion

Results for specific equivalent work and specific water cooling load for phase-changing absorbents and MEA are shown in Fig. 5. The figure shows the variability in the specific equivalent work and in the specific water cooling load within one standard deviation (±1 sigma) based on variation in the parameters defined in the Design of Experiment (DOE) for each capture technology and explained in sections 3.3 and 4.1. Results indicate that for all flue gas conditions studied in both retrofit and greenfield applications, the specific equivalent work necessary to operate phase-changing absorbents is expected to be lower than that for MEA capture.

Reduced energy requirements are expected for phase-changing absorbent compared to MEA because of three main factors: 1) less heat is required in the desorption process as a pure absorbent avoids the need for non-absorbing co-solvents that should be heated (water in the case of MEA), 2) phase change during absorption and desorption avoids the thermal separation and distillation processes needed in MEA capture and 3) lower CO_2 compression power is required as the desorption process in the phase-changing concept occurs at a pressure higher than MEA (~16 bar). This improved performance of phase-changing absorbents is expected to be more pronounced at higher CO_2 concentrations. While at 4 mole% CO_2 concentration in flue gas the specific equivalent work of phase-changing absorbent is 25% lower than MEA retrofit, at 13 mole% CO_2 the reduction could be as high as 42%. It is important to note that for all cases the specific equivalent work for MEA in greenfield applications is lower than in retrofit applications, as it is assumed that the conditions of extraction steam are more appropriate for the desorption process in greenfield cases than in retrofit cases.

With regard to the specific water cooling load, the lowest observable value occurs at 8 mole% CO_2 concentration for both phase-changing absorbents and MEA capture. At this CO_2 concentration, the specific water cooling load is 8% lower for MEA than for phase-changing absorbent. However, at 13 and 4 mole% CO_2 concentrations phase-changing absorbent presented 5% and 6% lower specific water cooling load than MEA respectively. In spite of offering potential lower specific equivalent work and specific water cooling load, the level of maturity of the phase-changing absorbent is certainly lower than MEA capture and therefore higher uncertainty in performance is expected. For instance the calculated standard deviation of the specific equivalent work in MEA was about 0,010 MJ/kg-CO_2, while for phase-changing absorbent was about 0,05 MJ/kg-CO_2.

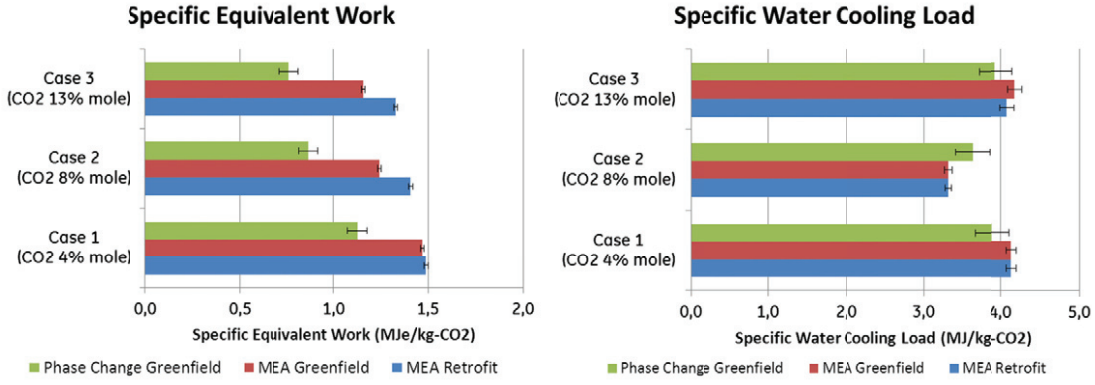


Fig. 5. (a) Comparison of specific equivalent work; (b) Comparison of specific water cooling load.

7. Conclusions

GE is actively investigating and developing novel technologies for both capturing and compressing CO2 from power plants with potential lower energy requirements and environmental impact less than state-of-the-art processes. This investigation compared the performance of phase-changing absorbents to state-of-the-art monoethanolamine (MEA) capture for different flue gas conditions in greenfield and retrofit applications.

Results indicate that depending on the flue gas conditions, the specific equivalent work necessary for operating phase-changing absorbents is expected to be up to 40% lower than for MEA capture. Besides lower energy requirements, a potential reduction of up to 6% in specific water cooling load might be expected for phase-changing absorbent over MEA, for the cases of 4 and 13 mole% CO2 concentrations. However, as the level of maturity of the alternative capture technology is certainly lower than MEA capture, higher uncertainty in performance is expected.

Acknowledgements

The information, data, or work presented herein was funded in part by the Advanced Research Projects Agency - Energy (ARPA-E), U.S. Department of Energy under the following contract: Award Number DE-AR0000084 in collaboration with University of Pittsburgh.

References

[1] DOE/NETL. Cost and performance baseline for fossil energy plants, Volume 1: bituminous coal and natural gas to electricity. Revision 2; 2010.

[2] Galindo Cifre P, Brechtel K, Hoch S, Garcia H, Asprion N, Hasse H, Scheffknecht G. Integration of a chemical process model in a power plant modeling tool for the simulation of an amine based CO_2 scrubber. *Fuel* 2009; **88**: 2481-2488.

[3] Korkmaz Ö, Oeljeklaus G, Görner K. Analysis of retrofitting coal-fired power plants with carbon dioxide capture. *Energy Procedia* 2009; **1**: 1289-1295

[4] Plaza J M, Van Wagener D, Rochelle G T. Modeling CO_2 capture with aqueous monoethanolamine. *International Journal of Greenhouse Gas Control* 2010; **4**: 161-166

[5] Kothandaraman A, Nord L, Bolland O, Herzog HJ, McRae GJ. Comparison of solvents for post-combustion capture of CO_2 by chemical absorption. *Energy Procedia* 2009; **1**: 1373-1380

[6] Oyenekan B A, Rochelle G T. Stripper models for CO_2 capture by aqueous solvents. *Greenhouse Gas Control Technologies* 2005; **7**: 1861-1864

[7] Berstad D, Arasto A, Jordal K, Haugen G. Parametric study and benchmarking of NGCC, coal and biomass power cycles integrated with MEA-based post-combustion CO2 capture. *Energy Procedia* 2011; **4**: 1737-1744

[8] Lawal A, Wang M, Stephenson P, Obi O. Demonstrating full-scale post-combustion CO2 capture for coal-fired power plants through dynamic modelling and simulation. *Fuel* 2010. Article in Press.

[9] Reddy S, Johnson D, Gilmartin J. Fluor's Econamine FG PlusSM technology for CO2 capture at coal-fired power plants. *Air and Waste Management Association - 7th Power Plant Air Pollutant Control 'Mega' Symposium 2008* 1, pp. 63-79

[10] Van Wagener D H, Rochelle G T. Stripper configurations for CO2 capture by aqueous monoethanolamine and piperazine. *Energy Procedia* 2011; **4**: 1323-1330

[11] Schach M O, Schneider R, Schramm H, Repke J U. Exergoeconomic Analysis of Post-Combustion CO2 Capture Processes. *Computer Aided Chemical Engineering* 2010; **28**: 997-1002

[12] Rochelle G T, et Al. CO_2 capture by absorption with potassium carbonate. *DOE Report, DE-FC26-02NT41440; 2007*